

Boost to Bioethanol Distillation by Internal Heat-Integrated Distillation Column (HIDiC)

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Abstract

A novel heat integrated distillation column (HIDiC) has been developed to enrich the yield of fermented bioethanol. The double-tube HIDiC column places the stripping section inside the inner tube and the rectifying section in the outer annular space. A dry vacuum pump placed between the sections makes the boiling temperature in the stripping section lower than that in the rectifying section, and the fermented mash (5 wt% ethanol) is concentrated to > 90 wt% ethanol near the azeotropic point at a low reflux ratio (0.378). When the stripping section pressure falls to 225 mm Hg while the rectifying section pressure is held at 1 atm, no live steam is required for the enriching process. Even if the preheater heat duty and the power requirement of the vacuum pump are considered, the depressurization of this HIDiC system achieves the energy-saving target (< 4 MJ/L-EtOH) of the NEDO project.

Keywords

HIDiC; Bio-ethanol; Energy Saving; Depressurized Distillation

Introduction

This research was conducted within the Consolidated Bio-Processing (CBP) governmental project, which was supported by the New-Energy and Industrial Technology Development Organization (NEDO) of Japan. The project set a total production cost goal for the production of 1 L of dehydrated ethanol as biofuel at JPY 40/L-EtOH (USD 0.4/L-EtOH). CBP also aimed at a 50% reduction of CO₂ emissions relative to gasoline. But reducing the cost of producing fuel ethanol depends not only on the enzyme consumption of fermentation from soft biomass but also on the energy consumption of the distillation process.

Distillation technology is useful for separation but does not permit a dramatic reduction of energy. With consideration of the distribution of energy required for

the overall CBP process, the target energy consumption was fixed at 4 MJ/L-EtOH. The aim of this research, therefore, is to dramatically reduce the energy consumption of the dehydration process by improving the heat-integrated distillation column (HIDiC) system, originally proposed by Mah et al. (1977) (Fig.1). A bench plant was designed to substantiate the results.

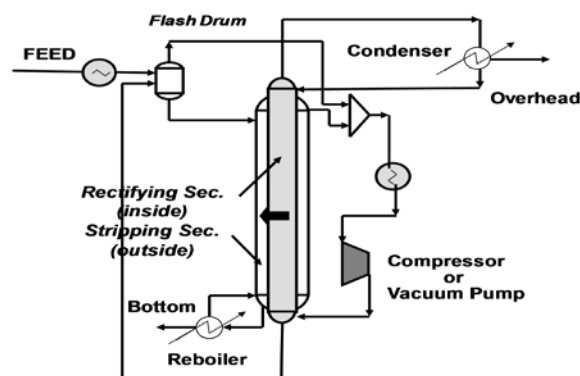


FIG. 1 CONFIGURATION OF DOUBLE-TUBE HIDiC SYSTEM

Background of This Research

The previous NEDO project, in which Kansai Chemical Engineering participated, saw the world's first success in developing and applying HIDiC technology to a C5-splitter in an oil refinery (Horiuchi et al. 2006, Kataoka et al. 2007). This project achieved a 60% reduction of heat energy consumed in the first column. However, HIDiC technology has not yet made inroads in the petroleum and petrochemical industry. It is of great importance to innovate HIDiC technology for the production of ethanol from biomass.

Internal heat integration based on the heat pump principle takes place as the transfer of heat from the

high-pressure rectifying section to the low-pressure stripping section. Partial condensation of upward vapour occurs at temperature Tr_i at the i th stage of the rectifying section, while partial vaporization of downward liquid occurs at temperature Ts_i at the i th stage of the stripping section. The overall heat transfer coefficient can be defined as shown in Fig.2. For convenience, the heat transfer area A_i is defined as the surface area of the partition wall corresponding to one tray-to-tray spacing. It is very difficult to estimate the area of the partition wall that causes vapour condensation in the rectifying section and liquid vaporization in the stripping section.

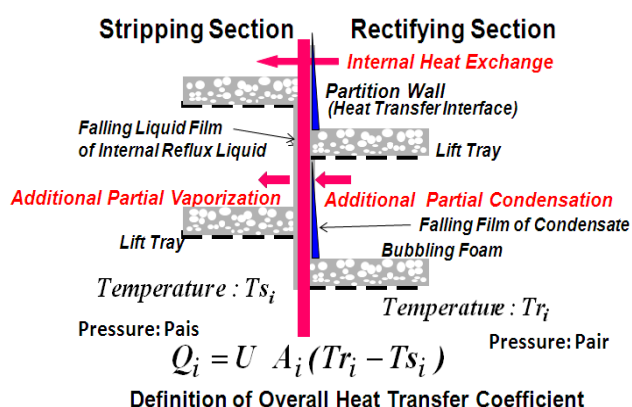


FIG. 2 SITUATION OF INTERNAL HEAT EXCHANGE AND DEFINITION OF OVERALL HEAT TRANSFER COEFFICIENT

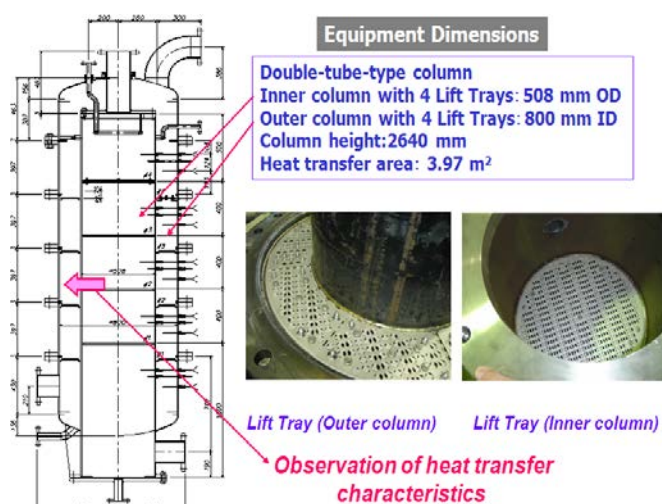


FIG. 3 EXPERIMENTAL DOUBLE-TUBE HIDiC COLUMN

An experimental HIDiC apparatus shown in Fig. 3 (e.g. Noda et al. 2009) was used to create a database for HIDiC plant design, which recorded the overall heat transfer coefficient, temperature difference and plate efficiency for several binary systems (e.g. benzene- toluene and ethanol- water).

The diameter of the double-tube column was initially designed for the commercial scale, to deal with 1.6 t/h

of benzene-toluene binary mixture. The tray-to-tray spacing was 300 mm. The HIDiC control parameter is a compression ratio, defined as the ratio of rectifying section pressure P_{air} to the stripping section pressure P_{ais} .

The inner column serves as the rectifying section and the outer annular space as the stripping section. Both sections are equipped with our original dual-flow trays "Lift Trays".

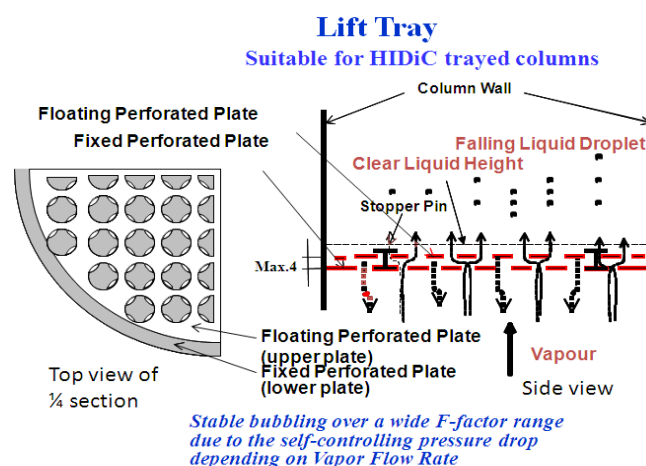


FIG.4 LIFT TRAY CONSISTING OF TWO PERFORATED PLATES

The Lift Trays control the tray pressure drop automatically: increased vapour flow lifts the upper perforated plate and thus increases the opening area (Fig.4). The wide range of stable F-factor values makes this self-controlling tray suitable for HIDiC systems, because the internal heat exchange causes large variations in vapour and liquid rates within the column: owing to the internal heat exchange, the vapour and liquid flow rates become minimal near the bottom and maximal near the top in the stripping section and vice versa in the rectifying section. The experiment was conducted with two modes of operating pressure: (1) pressurizing mode, raising the rectifying section pressure to $P_{air} > 1$ atm while the stripping section was held at 1 atm; and (2) depressurizing mode, lowering the stripping section pressure to $P_{ais} < 1$ atm, while the rectifying section was held at 1 atm. Experimental data for an ethanol-water system are used here to explain this bench plant design.

Figure 5 shows the temperature difference $\Delta T = Tr - Ts$ between the rectifying and stripping section averaged over the four stages of the experimental apparatus. The linear relation with the compression ratio results from the fact that the boiling point rises with pressure.

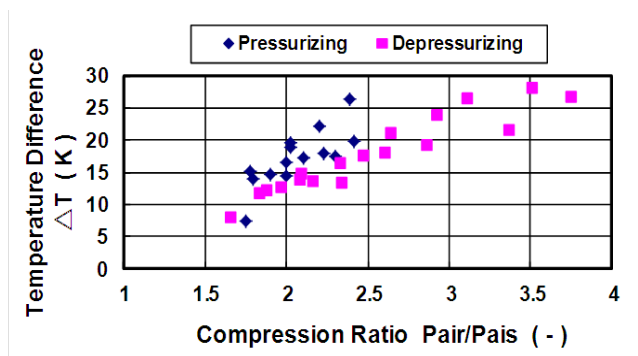


FIG. 5 VARIATION IN TEMPERATURE DIFFERENCE WITH COMPRESSION RATIO.

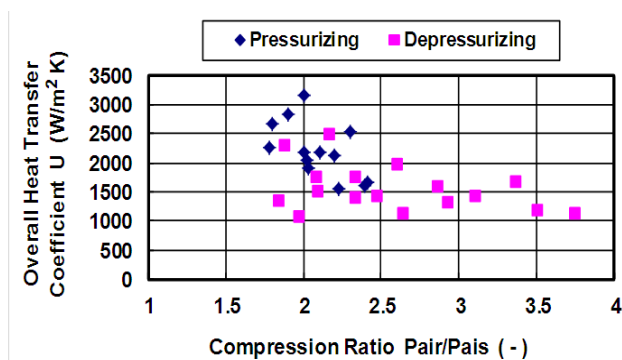


FIG. 6 EXPERIMENTAL OVERALL HEAT TRANSFER COEFFICIENT

For the depressurizing mode, a very low pressure is required in the stripping section to give a large temperature difference. Figure 6 shows how the overall heat transfer coefficient U varies with the compression ratio for the ethanol-water system. For the depressurizing mode, the heat transfer coefficient tends to diminish gradually with compression ratio, because the thickness of the condensate film on the rectifying side increases as the temperature difference increases. In this experimental condition using a clean and clear aqueous solution of pure ethanol, relatively high heat transfer coefficients were obtained because there was no fouling. This suggests that in the design of a bench plant it is necessary to underestimate the overall heat transfer coefficient in the process simulation analysis as the fouling effect is estimated on account of the fermentation residues. As explained above, the heat transfer area was determined as the cylindrical wall surface of the inner column corresponding to the tray-to-tray spacing. This suggests that the plate efficiency of the Lift Tray and the tray-to-tray spacing must be specified in the evaluation of internal heat exchange, even for a simulation analysis assuming an ideal plate condition. The number of ideal plates was computed by the McCabe-Thiele step-by-step method, taking into account the varying liquid and vapour flow rates

experimentally observed in the condition of internal heat exchange (Fig.7). The plate efficiency is defined as the ratio of the number of ideal plates to the number of Lift Trays of the experimental apparatus.

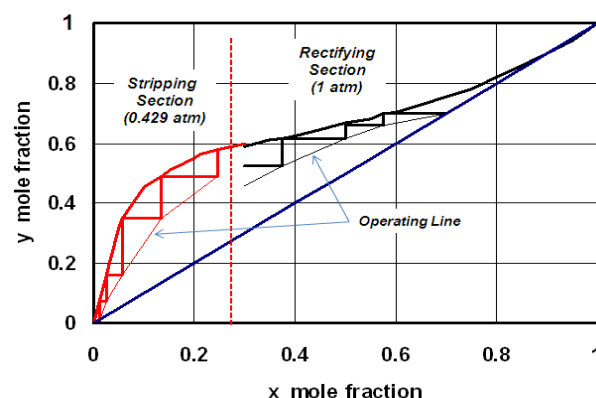


FIG. 7 STEP-BY-STEP CALCULATION OF THE NUMBER OF IDEAL PLATES IN THE HIDIC CONDITION

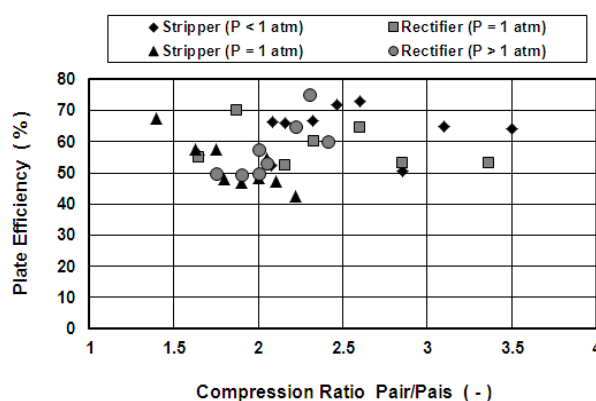


FIG. 8 EXPERIMENTAL DATA OF LIFT TRAY PLATE EFFICIENCY

Figure 8 shows the plate efficiency data obtained with the experimental apparatus. Under the experimental conditions, most plate efficiency points lie above 50%.

Preliminary Test of Fouling by Fermented Mash

The most important point is how to prevent the fermentation residues (e.g. dissolved saccharides) from depositing onto the heat transfer surface of the stripping section and reboiler. After filtration to remove solid suspensions, the effect of boiling point using mash fermented from rice straw was investigated in a jacketed tank evaporator. The deposition of solidified substances on the inside heat surface of the evaporator was observed as the operating pressure, (i.e. the boiling point) was varied.

The photograph on the left of Fig. 9 indicates that serious fouling occurred on the inner heat transfer surface of the evaporator under normal-pressure boiling. On the other hand, the photograph on the right suggests that the solidification reaction (e.g. due

to glucans) can be suppressed if the boiling point is kept at 68.2°C by lowering the pressure to 235 mm Hg.



FIG. 9 TEST OF EVAPORATOR FOULING (WHEN $p = 235$ MMHG, THE BOILING POINT WAS 68.2°C)

The depressurizing mode should be used to prevent the stripping section from becoming fouled. Steam distillation to supply live steam should be used instead of a bottom reboiler. In addition, for easy cleaning and maintenance, an actual double-tube HIDiC plant should have the stripping section placed in the inner column.

Simulation Analysis for Design of HIDiC Bench Plant

A simulation analysis was done to determine the design specification of the HIDiC bench plant.

The main purpose of this project is to use the HIDiC bench plant to enrich the overhead product from 5 wt% ethanol in fermented mash (50 L/h = 49.35 kg/h) to >90 wt% with a recovery of >95% of the ethanol in the feed. The target for the total energy consumed by the bench plant in producing 1 L of ethanol is 4 MJ/L-EtOH.

Table 1 shows the conditions and specifications used for the simulation analysis.

TABLE 1 CONDITIONS AND SPECIFICATIONS FOR SIMULATION ANALYSIS

Column	Double-tube type
Stripping section (inner column)	150 mm ID, 6.8 m
Lift Tray	34 plates, (17 ideal plates)
Plate efficiency(%)	50 (assumed)
Pressure, Pais, (column top, mmHg)	variable
Pressure drop per tray (mmHg)	6.4 (assumed)
Plate-to-plate spacing (mm) (actual plate)	200
Feed Plate (ideal plate basis)	#2
Rectifier bottom liquid supply	#1
Rectifying section (outer column)	250 mmID, 5.2 m
Structured packing	13 ideal plates
Pressure, Pair, (column top, mmHg)	740
Pressure drop in total (mmHg)	10

Reflux ratio	variable
Distillation condition	
Feed rate (L/h) initial temperature: 30°C	50 (= 49.35 kg/h)
ethanol concentration (wt%)	5.0
feed temperature :	69°C
Overhead product	
ethanol concentration (wt%)	>90
ethanol recovery (%)	>95
Bottom product	
ethanol concentration (wt%)	<0.1
Heat duty (Live steam instead of reboiler)	
Live steam (0.3 MPa) supply (kg/h)	variable
Preheater heat duty (live steam supply)	variable
Cooling duty (condenser), boiling temperature, 740 mmHg	variable
Internal heat exchange	
Overall heat transfer coefficient (W/m ² K) x heat transfer area (m ²) of one stage-to-stage spacing, UA (W/K) (ideal stage basis)	54.82
Number of active stages (Lift tray basis)	25 (#3 ~ #27)
Heat loss assumed	none
Dry vacuum pump	variable
Theoretical power consumption (kW)	

The simulation requires the actual heat transfer area for the internal heat exchange calculation. The number of ideal stages in the stripping section is determined by assuming a Lift Tray plate efficiency of 50% (Fig. 8). This implies that each ideal stage should have 400 mm as the stage-to-stage spacing, i.e. twice the tray-to-tray spacing of actual trays (= 200 mm). Therefore, one ideal stage should have twice the heat transfer area of one actual tray-to-tray spacing. The height of the packing bed of the rectifying section with 13 ideal stages is set to the height of the internal heat exchange tray section of the stripping section.

PRO/II by SimSci-Esscor (Invensys) was used as the process simulation program, and NRTL as the vapour-liquid equilibrium model.

Figure 10 shows a schematic picture of the mass and enthalpy balances on the i th plate of the rectifying and stripping sections. The following set of governing equations was used in the simulation analysis.

(Stripping section)

$$\begin{aligned}
 V_{i+1} + L_{i-1} &= V_i + L_i \\
 V_{i+1}y_{i+1} + L_{i-1}x_{i-1} &= V_iy_i + L_ix_i \\
 V_{i+1}H_{V,i+1} + L_{i-1}H_{L,i-1} + Q_{\text{int},i} &= V_iH_{V,i} + L_iH_{L,i} \\
 Q_{\text{int},i} &= U A_i (T_{r,i} - T_{s,i}) \\
 V_i - V_{i+1} &= Q_{\text{int},i} / \lambda_{s,i}
 \end{aligned}$$

(Rectifying section)

$$\begin{aligned}
 V_{i+1} + L_{i-1} &= V_i + L_i \\
 V_{i+1}y_{i+1} + L_{i-1}x_{i-1} &= V_iy_i + L_ix_i \\
 V_{i+1}H_{V,i+1} + L_{i-1}H_{L,i-1} - Q_{int,i} &= V_iH_{V,i} + L_iH_{L,i} \\
 Q_{int,i} &= U A_i(T_{r,i} - T_{s,i}) \\
 V_i - V_{i+1} &= -Q_{int,i} / \lambda_{r,i}
 \end{aligned}$$

(Dry vacuum pump)

$$Pw_{theo} = V_1(E_{V1out} - E_{V1in})$$

(Live steam)

$STM = (W_1E_{W1} + V_1E_{V1} - W_2E_{W2} - FE_F) / E_{STM}$ where V_i, L_i are the flowrates (kmol/h) of vapour and liquid leaving the i th stage, x_i, y_i are the ethanol concentrations (mole fractions) of vapour and liquid, $H_{V,i}, H_{L,i}$ are enthalpies (kJ/kmol) of vapour and liquid on the i th stage, and $\lambda_{s,i}, \lambda_{r,i}$ are latent energies (kJ/kmol) of vaporization and condensation on the i th stage. The equation evaluating $V_i - V_{i+1}$ of each section indicates an increase in vapour flow rate due to additional vaporization (stripping section) or a decrease due to additional condensation (rectifying section).

The internal heat exchange is calculated at each stage from the temperature difference $T_{r,i} - T_{s,i}$ between the rectifying and stripping sections at the i th plate.

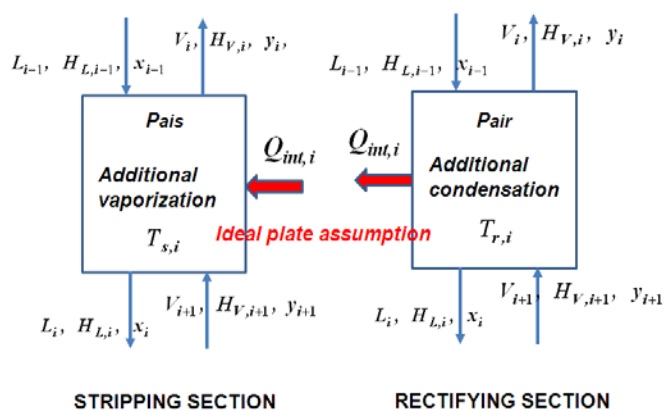


FIG.10 SCHEMATIC PICTURE OF MASS AND ENTHALPY BALANCES AT ITH STAGE

The theoretical power requirement Pw_{theo} of the dry vacuum pump is calculated by using the enthalpy change $E_{V1out} - E_{V1in}$ of vapour from the inlet to the outlet of the pump due to an assumed isentropic compression. The rate of live steam supply, STM , is calculated by using the enthalpy balance applied to the stripping section: flow rates W_1, W_2, V_1, F are, respectively, bottom product of the stripping section, bottom product of the rectifying section, overhead

vapour from the stripping section and feed.

The inner diameters of the stripping and rectifying sections were determined from the vapour and liquid flow rates obtained by the preliminary simulation analysis so that the internal vapour velocity F-factor is $F < 0.5 (m/s)(kg/m^3)^{0.5}$ in both sections,

when the diameters of the rectifying and stripping sections are respectively $D_o = 250$ mm and $D_i = 150$ mm.

The area of heat transfer due to partial condensation on the rectifying side should depend mainly on the bare wall not touched by the bubbling foam, whereas the area of heat transfer for partial vaporization on the stripping side should be controlled by the wall wetted with the internal reflux liquid. For engineering purposes in plant design, it is therefore not expedient to use the effective heat transfer area for evaluating U . For convenience, the cylindrical wall surface of the inner column cut by one ideal stage-to-stage spacing is used as the heat transfer surface area: $A_i = 0.1885$ m². For the case of dirty mash liquid, there is a high possibility that the practical overall heat transfer coefficient may drop by half compared with the experimental data (Fig.6) owing to dissolved saccharides. For practical engineering safety, therefore, the heat transfer coefficient is assumed to be kept constant at $U = 290.8$ W/m²K (= 250 kcal/m²hr°C) over the column height with internal heat exchange. No heat loss from the outer surface of the HiDiC column is assumed for the analysis. To avoid the fouling problem due to the solidification of mash residues, the stripping section is depressurized to <1 atm and live steam (0.3 MPa) is supplied at the bottom instead of reboiler heat duty.

The simulation analysis gives the variation in vapour velocity expressed as the F-factor by column height (Fig. 11). The vertical axis indicates the ideal stage number counted from the bottom. The ideal stage-to-stage spacing is assumed to be the same (400 mm) for calculation of internal heat exchange in both the sections. Thirteen ideal stages (#1~#13) of the entire rectifying section is made coincident with the thirteen ideal stages (#2~#14) of the internal heat exchange region of the stripping section.

The vapour velocity decreases upward in the rectifying section owing to additional condensation, but increases upward in the stripping section owing to additional vaporization due to the latent heat released from the rectifier side.

The F-factors are made very small (1) to suppress the effect of entrainment from the bubbling foam to the above trays and (2) to ensure a large enough area of heat transfer from the rectifying to the stripping sections by diminishing the depth of the bubbling foam layer. In other words, the diameter of the inner column is made larger than is needed to make the F-factor small.. The internal reflux liquid velocity varies similarly (Fig. 12).

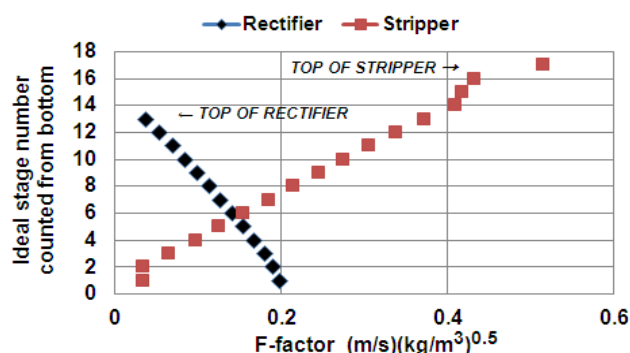


FIG.11 VERTICAL VARIATION IN VAPOUR VELOCITY

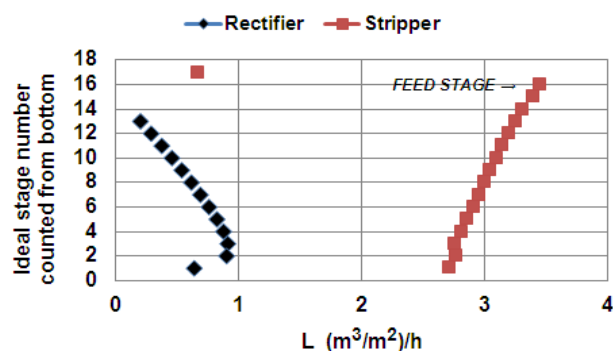


FIG.12 VERTICAL VARIATION IN LIQUID VELOCITY

The liquid velocity at the top (ideal stage #17) of the stripping section is very small because the feed plate is ideal stage #16. The vapour rate becomes very small near the bottom of the stripping section and near the top of the rectifying section. As the vapour and liquid have very large variation in the vertical direction, perforation of the Lift Trays was designed carefully.

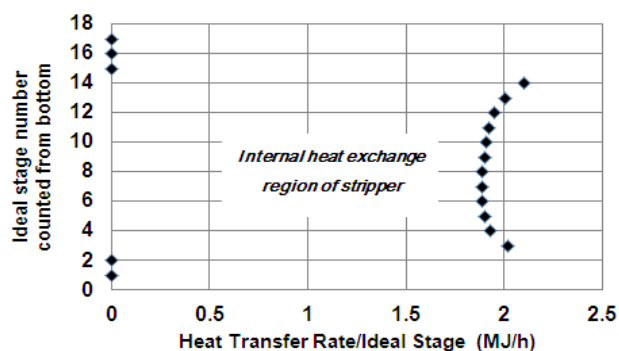


FIG.13 VERTICAL VARIATION OF HEAT TRANSFER RATE PER IDEAL STAGE

Figure 13 shows how the internal heat exchange rate per ideal stage varies in the vertical direction. For the simulation analysis, the heat transfer coefficient was assumed to be constant at $U = 290.8 \text{ W/m}^2\text{K}$. Therefore, the temperature difference between the rectifying and stripping sections varies similar to heat transfer rate.

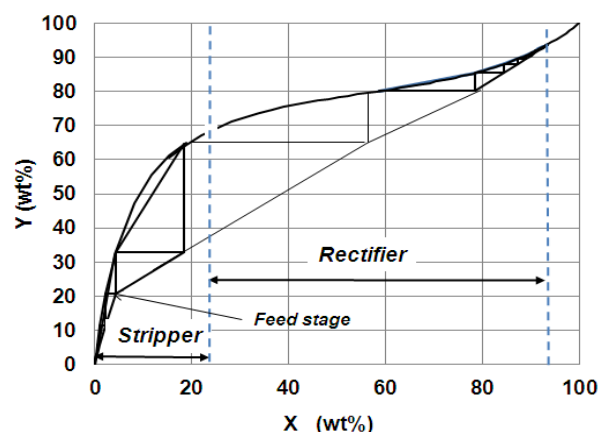


FIG.14 VARIATIONS IN VAPOUR AND LIQUID CONCENTRATIONS BY MCCABE-THIELE METHOD

Figure 14 shows the variations in vapour and liquid concentrations in the form of a McCabe-Thiele step-by-step diagram when the stripping section is depressurized to $P_{ais} = 210 \text{ mm Hg}$ while the external reflux ratio is set as 1.0. Strictly speaking, this distillation process accompanied by internal heat integration cannot be expressed by the McCabe-Thiele diagram. The slope of the operating line varies slightly owing to additional condensation and vaporization due to internal heat integration. The liquid from the rectifier bottom is supplied to the top of the stripper. The steep operating line near $x = 5 \text{ wt\%}$ implies that the feed stage is located at #2 ideal stage from the top.

Figure 15 shows the simulation results when the stripping section pressure is fixed at 210 mm Hg. These values were used as the bench plant design specification. According to the results, the energy consumption comprises the preheater heat duty, the live steam supply rate instead of the reboiler heat duty, and the electricity requirement of the vacuum pump. The analysis gives a theoretical power requirement of 0.70 kW for the case of isentropic change.

If the practical efficiency of a vacuum pump is assumed to be 40%, its power requirement is

$$E_{VP} = (0.70/0.40)(3600/1000) = 6.3 \text{ MJ/h}$$

The total energy consumption theoretically evaluated by the simulation is given by

$$Q_{total} = Q_{STM} + E_{VP} + Q_{preh}$$

$$= 2.15 + 6.3 + 2.35 = 10.8 \text{ MJ/h.}$$

The volumetric rate of ethanol contained in the feed is

$$V_{EtOH} = (49.35)(0.05)/0.78 = 3.16 \text{ L/h}$$

The energy required for producing 1 L of ethanol is

$$Q_{total} / V_{EtOH} = 10.8/3.16 = 3.42 \text{ MJ/L-EtOH.}$$

This value suggests a good chance of attaining the NEDO project target (< 4 MJ/L-EtOH).

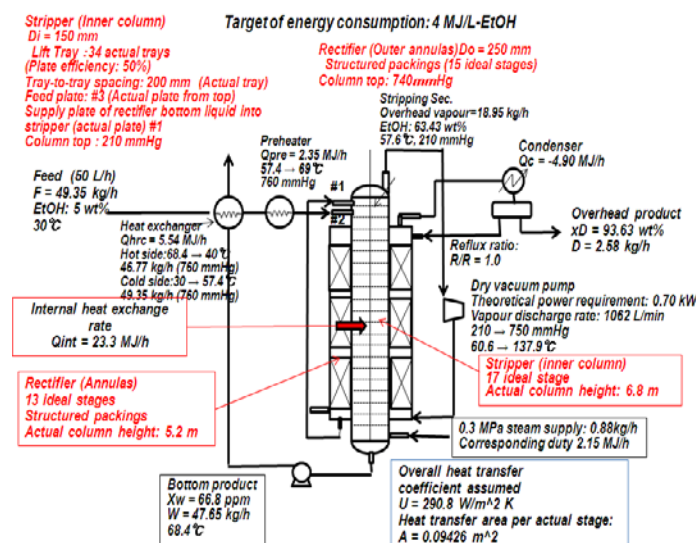


FIG. 15 BENCH PLANT DESIGN SPECIFICATION OBTAINED BY PROCESS SIMULATION ANALYSIS

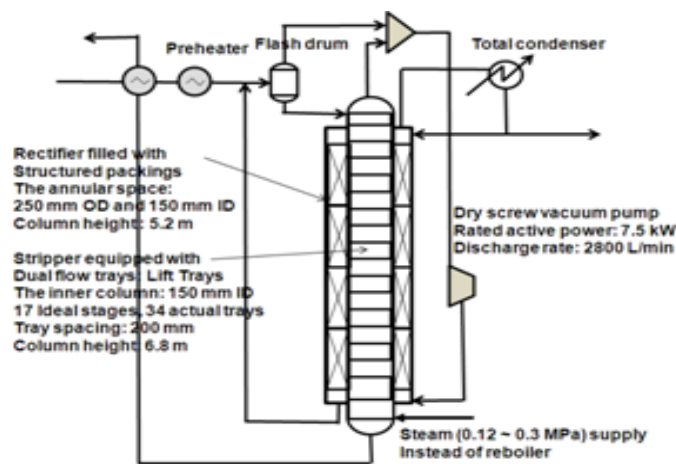


FIG. 16 DOUBLE-TUBE HIDiC BENCH PLANT

HIDiC Bench Plant

On the basis of the simulation results, a HIDiC bench plant was designed for producing an overhead product of >90 wt% ethanol from 50 kg/h of fermented mash (5 wt% ethanol). The double-tube HIDiC column has an inner column (150 mm ID, 6.8 m in height) as the stripping section and an outer annular

column (250 mm ID, 5.2 m in height) as the rectifying section (Fig. 16.)

The stripping section is equipped with dual-flow Lift Trays inside with no downcomer (downflow pipe). As distinct from the standard HIDiC configuration, the stripping section is placed in the inner column only for easy washing of the fermentation residues. To avoid the solidification and deposition of mash residues, a screw-type dry vacuum pump is necessary to keep the pressure (i.e. bubbling point) of the stripping section lower than 250 mm Hg.

A screw-type dry vacuum pump (2800 L/min, 7.5 kW) was installed between the rectifying and stripping sections to keep the stripping section pressure at < 250 mm Hg while the rectifying section pressure was normal. The cylindrical wall of the inner column serves as the heat transfer surface for internal heat integration. The rectifying section is filled with structured packing because the vapour and liquid are clean. Engineering experience gained with Lift Trays in various distillation processes suggests that the height of the structured packing corresponding to one ideal stage can usually be assumed to be less than twice the tray-to-tray spacing of the Lift Tray. For internal heat exchange only, however, the height of the rectifying section filled with structured packing should be set to the height of the heat transfer part with 25 Lift Trays (#3 ~ #27) in the stripping section. To avoid fouling, the bottom of the stripping section has a nozzle of live steam supply instead of a bottom reboiler. On the assumption of 50% plate efficiency, the stripping section has 32 Lift Trays inside, and the rectifying section is filled with structured packing (corresponding to 13 ideal plates). The heat transfer area is evaluated from the actual tray-to-tray spacing of 200 mm. After filtration, the clear fermented mash liquid is fed through the preheater into actual tray #3 of the stripping section. The vapour from the top of the stripping section is pressurized and supplied to the bottom of the rectifying section by means of the dry vacuum pump. The liquid from the bottom of the rectifying section is supplied to actual tray #1 (tp) of the stripping section through the flash drum.

Result and Discussion of Test Operation

Operating Conditions

The simulation result (Fig.17) indicates that a feed concentration of >3 wt% ethanol allows the energy consumption target of <4 MJ/L-EtOH to be met.

Allowing for actual heat loss, the fermentation target of the project was fixed as 5 wt% ethanol for optimal energy-saving in the enriching process.

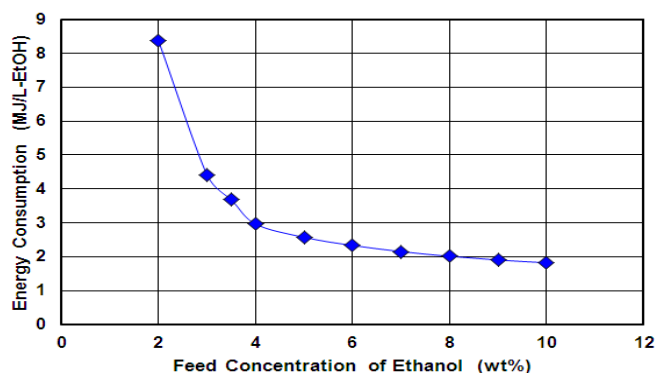


FIG.17 ENERGY CONSUMPTION VS. FEED CONCENTRATION

Result of Test Operation

Feed liquid to simulate fermented mash was prepared because the amount of fermented mash produced by the CBP process was not enough for a long run of the bench plant. Therefore the effect of fouling was not able to be investigated by a long run of test operation. The simulation analysis was done how to raise the compression ratio by a vacuum pump to ensure the rate of internal heat transfer but its result still remains to be checked by the actual operation.

Holding the stripping section at 225 mm Hg gave the best results (Fig. 18).

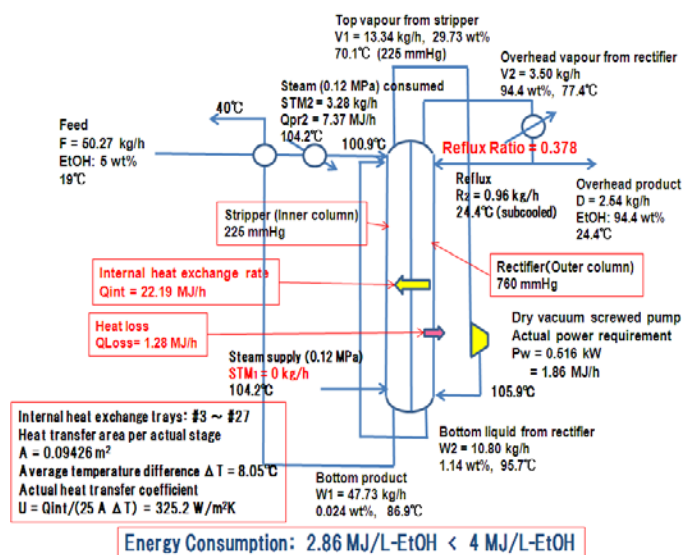


FIG.18 SUMMARY OF THE BEST OPERATION OF THE BENCH PLANT

This bench plant successfully recovered 95.4 % of the ethanol in the feed, gave an ethanol concentration in the overhead product of 94.4 wt%, and left only 0.024 wt% ethanol in the bottom from the stripping section. This result also satisfied the specified effluent

condition of < 0.1 wt% EtOH.

The stripping section was placed in the inner column of the bench plant but in the outer annular space for the simulation analysis.

In addition, the simulation analysis (Fig. 15) assumed that the column top of the stripping section was set at 210 mm Hg, while the test operation of the bench plant gave a satisfactory result (Fig. 18) when it was 225 mm Hg. Therefore, it was a little bit difficult to make a strict comparison between the simulation analysis and the actual test operation. The test operation gave a good result when the rate of live steam supply was zero, but the simulation analysis required 0.88 kg/h. This value corresponded to a heat duty of 2.15 MJ/h, which was based on the latent heat of steam of 0.3 MPa. The simulation analysis assumed no heat loss, while the bench plant had a heat loss of 1.28 MJ/h. The difference in the heat duty was small in comparison with the rate of internal heat exchange. The simulation analysis and the test operation gave almost equal rates of internal heat exchange 23.3 MJ/h (simulation analysis) and 22.19 MJ/h (test operation). In the context of biofuel, the project did not need to take into account fermentation by-products such as acetic acid, formic acid and furfural, which were discharged mainly from the bottom of the stripping section.

Energy Consumption

The best result was obtained with the reflux ratio of 0.378 when no live steam was required as the heat duty for the stripping section: $Q_{STM} = 0$ MJ/h. This means that no heat duty is required for the distillation column, except for the energy to run the dry vacuum pump. The actual electricity requirement of the dry vacuum pump was measured as $E_{VP} = 0.516$ kW = 1.86 MJ/h. This value is less than the theoretical power requirement of 0.70 kW (Fig. 15), probably owing to the difference in the stripping section pressure.

The cooling duty of the overhead condenser was not taken into consideration, because town water was used as cooling water. In addition, the power requirement of various pumps for liquid transportation is negligibly small in the discussion of total energy use.

The rate of live steam (0.12 MPa) required for preheating the feed was 3.58 kg/h. This corresponded to a heat duty of $Q_{preh} = 7.37$ MJ/h.

Therefore, the total energy required for this test

operation of the HIDiC bench plant is

$$Q_{total} = Q_{STM} + E_{VP} + Q_{preh} = 9.23 \text{ MJ/h}$$

The volumetric flow rate of ethanol included in the feed of 50.27 kg/h is $V_{EtOH} = 3.22 \text{ L/h}$ at 30°C .

The energy required for obtaining 1 L of ethanol as distillate by the HIDiC bench plant is

$$Q_{total} / V_{EtOH} = 2.87 \text{ MJ} / \text{L} - \text{EtOH}$$

This value is less than the project target of 4 MJ/L-EtOH.

The estimate suggests a saving of approximately 60% relative to the energy consumption of an ordinary distillation column calculated in the same distillation conditions.

If the pressure at the top of the stripping section is reduced below 225 mm Hg, the live steam rate required for the preheater will be decreased, offering the possibility of further improving the energy savings.

Characteristics of Internal Heat Exchange

The internal heat transfer equation (Fig. 2) can be rewritten as

$$U = \frac{Q_{int}}{A \Delta T}$$

where Q_{int} is the total rate of heat transfer from the rectifying to stripping section, A is the total heat transfer surface area (i.e. the cylindrical wall surface of the heat transfer part of the stripping section) and ΔT is the average temperature difference between the rectifying and stripping sections

Internal heat exchange takes place in the region of actual trays #3 to #27 (25 in total). One tray-to-tray spacing (= 200 mm) of the Lift Tray has an internal heat transfer area of 0.09426 m^2 . On the rectifying side, partial condensation occurs on the outer surface of the inner column; on the stripping side, partial vaporization takes place on the inner surface of the inner column, owing to the heat input from the rectifying section. For simplicity, the entire cylindrical wall surface enclosed by each tray-to-tray spacing was assumed to be effective.

Various saccharides in the mash liquid might decrease the overall heat transfer coefficient. Therefore, the HIDiC bench plant was designed on the assumption of an overall heat transfer coefficient of $290.8 \text{ W/m}^2\text{K}$ (=

$250 \text{ kcal/m}^2\text{h}^\circ\text{C}$). This was an underestimate compared with the experimental data in our database (Fig. 6).

For the actual test operation, the average temperature difference between the rectifying and stripping sections was $\Delta T = 8.05^\circ\text{C}$.

The enthalpy balance calculated using the measured flow rates, temperatures and concentrations gave the rate of internal heat exchange between the rectifying and stripping sections as $Q_{int} = 22.19 \text{ MJ/h}$. (The simulation analysis gave 23.3 MJ/h .) The total energy consumption was the sum of the live steam supply to the stripping section, the vacuum pump power requirement and the live steam consumed by the preheater: $Q_{total} = Q_{STM} + E_{VP} + Q_{preh} = 9.23 \text{ MJ/h}$.

If $Q_{total} + Q_{int} = 31.42 \text{ MJ/h}$ is regarded as the total energy necessary for this distillation process, the internal heat exchange will compensate for 70% of the total. This implies a great energy-saving effect.

The total heat transfer surface area is given by $A = (0.09426 \text{ m}^2/\text{stage}) \times 25 \text{ stages} = 2.357 \text{ m}^2$. The actual overall heat transfer coefficient is given as

$$U = \frac{Q_{int}}{A \Delta T} = \frac{(22.19 / 3600)(10^6)}{(2.357)(8.05)} = 324.9 \text{ W/m}^2\text{K}$$

This value is only 11.7% larger than the design specification. As a result of enthalpy balance, the heat loss from the outer surface of the rectifying section is

$$Q_{Loss} = 1.28 \text{ MJ/h}$$

This value is small compared with the internal heat exchange rate of 22.19 MJ/h .

After the successful test operation, the bench plant (Fig. 19) has been delivered as the outcome of the NEDO project (Fig. 20). The internal structure is shown in Fig.19. The stripping section is placed inside for maintenance of washing out the mash residues. The rectifying section is filled with structured packings.

Figure 20 is a photograph of the HIDiC bench plant (the left side column) under installation. The other column (on the right side) is for another purpose.

Conclusions

A new HIDiC bench plant has been developed to enrich the distillation of cellulose ethanol produced from soft biomass at an energy saving of 2.86 MJ/L-EtOH , lower than the project target of 4 MJ/L-EtOH .

It improved performance to >90 wt% of ethanol. It can be regarded as the world's first successful application of a depressurized HiDiC..

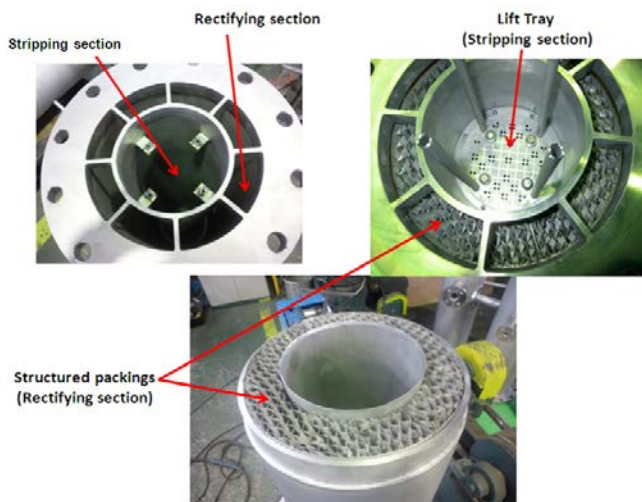


FIG.19 INTERNAL STRUCTURE OF BENCH PLANT. THE STRIPPING SECTION IS PLACED INSIDE FOR EASE OF WASHING OUT THE MASH RESIDUES. THE RECTIFYING SECTION IS FILLED WITH STRUCTURED PACKING.



FIG.20 HiDiC BENCH PLANT UNDER INSTALLATION (LEFT). (THE COLUMN ON THE RIGHT HAS ANOTHER PURPOSE.)

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